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Air Pollution Control Device Configurations

Acurex Corp.
Mountain View, CA

Prepared for

California Energy Commission
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CONSULTANT REPORT

AIR POLLUTION CONTROL DEVICE CONFIGURATIONS

MARCH 1980

P300-80-003

AIR POLLUTION CONTROL DEVICE CONFIGURATIONS

Acurex Corporation

California Energy Commission

Contract 300-039(8/9)

Project Officer, Don Kondoleon

March 1980

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Air Pollution Control Device Configurations

ABSTRACT

This report presents an analysis of those candidate air pollutant abatement schemes identified previously by Acurex Corporation in the report "Environmental Control Technologies for a Northern California Coal-Fired Power Plant" (300-039) for possible use on PG&E's proposed Montezuma 1 and 2 (previously called Fossil 1 and 2) electric generating plants. Emphasis is placed on the effects of the arrangement of the control components, namely the baghouse, selective catalytic reduction (SCR) and flue gas desulfurization (FGD) units.

Three basic configurations were identified as workable arrangements. These configurations are as follows:

- Arrangement A --

  ESP --- SCR --- Baghouse --- FGD --- Stack

- Arrangement B --

  Baghouse --- SCR --- FGD --- Stack

- Arrangement C --

  Baghouse --- FGD --- SCR --- Stack

The configurations were analyzed with respect to their capital, operating and maintenance cost impacts, reliability implications, and overall system removal efficiencies. Since there is no actual experience reported for these configurations with coal-fired boilers, reliability and overall efficiency estimates were described qualitatively.

Air Pollution Control Device Configurations

CEC PERSPECTIVE

This report examines possible emissions control equipment arrangements as might be employed on the coal-fired Montezuma 1 and 2 power plant (previously called Fossil 1 and 2), a 1600 MWe facility proposed by the Pacific Gas and Electric Company. The work reported herein was performed by Acurex Corporation and supplements an earlier investigation by Acurex, entitled "Environmental Control Technologies for a Northern California Coal-Fired Power Plant" (300-039).

Emission controls consisting of a selective catalytic reduction (SCR) unit, baghouse, and flue gas desulfurization (FGD) unit were alternatively configured as follows:

- Arrangement A

  ESP --> SCR --> Baghouse --> FGD --> Stack

- Arrangement B

  Baghouse --> SCR --> FGD --> Stack

- Arrangement C

  Baghouse --> FGD --> SCR --> Stack

The configurations were analyzed with respect to their capital, operating, and maintenance cost impacts, reliability implications, and overall system removal efficiencies. Since there is no actual experience reported for these configurations with coal-fired boilers, reliability and overall efficiency were described qualitatively.

Design Considerations

The primary attractiveness of Arrangement A is that flue gas leaving the boiler's

economizer is at the same temperature as the required operating temperature of the SCR process. However, the relatively high particulate grain loading of the flue gas may adversely effect the SCR catalyst bed. To alleviate this potential problem, a low efficiency hot electrostatic precipitator has been included to remove the bulk of the particulate effluents.

The formation of ammonium bisulfate and ammonium sulfate is possiblw with the use of injected NH_3 in the SCR process if there is any unreacted NH_3 available downstream. These compounds may foul downstream hardware and also create potential environmental problems. The careful monitoring of ammonia to ideal stoichiometric ratios will minimize bisulfate and sulfate formation.

Flue gas in Arrangement A enters the scrubber following particulate removal by the baghouse. With virtually no flyash, the problem of wet-dry interface disposition of solids is alleviated. Mist eliminators are designed to remove condensed water and slurry carryover from the FGD scrubber.

In contrast to Arrangement A, Arrangement B requires three heat transfer steps. This arrangement is more energy intensive than Arrangement A as the various heaters require a heat fluid for the recovery of energy from the SCR effluent stream and cycle steam or auxiliary fossil fuel for two reheaters.

Compared to Arrangement A, the baghouse placement in Arrangement B is advantageous. Upstream of the SCR process, baghouse performance is not affected by the formation of ammonium sulfate/bisulfate compounds and the potential adherence to the bag surface. Moreover, the effects of particulate erosion and blinding in the SCR reactor catalyst bed are minimized with this configuration.

In Arrangement C, the temperature of the flue gas leaving the boiler's economizer is reduced in an air preheater, followed by particulate control in the baghouse.

The gas then proceeds to the FGD system for sulfur dioxide removal. Flue gas scrubbed of SO_2 leaves the FGD system and enters a heat exchanger which utilizes recirculated flue gas from the SCR to heat the flue gas. An additional heater elevates the flue gas temperature to 750°F, prior to entering the SCR system.

Problems associated with flyash poisoning the SCR reactor have been essentially eliminated by the upstream baghouse and FGD process. However, particulates escaping the mist eliminator may interfere with SCR reactor operation due to the latter's "sticky" characteristics.

Cost

Arrangement A has the highest capital cost (within the accuracy of the cost estimates) of the three arrangements by 18 percent at $200 per kilowatt, due to the need for an additional major unit (electrostatic precipitator). However, operating and maintenance costs are lower by over 32 percent at 6.7 mills per kilowatt-hour in 1985. Arrangement B and C have lower capital costs as no ESP is required. The higher operating and maintenance costs for these arrangements result from the additional flue gas reheating. The capital cost of Arrangements B and C are similar at $170 per kilowatt. Therefore, the total annualized cost in mills per kilowatt-hour in 1985 for Arrangement A is 39.4, for Arrangement B is 39.2 and for Arrangement C is 38.4.

Reliability

Arrangement A requires two devices (SCR and hot ESP) that are in the development stage whereas Arrangements B and C require only one (SCR). The newly developed technologies are not expected to be as reliable as currently available equipment. Other reliability problems, such as heat exchanger corrosion, can be accommodated though design (at higher costs) to assure high reliability. Arrangement A should

be considered, at least initially, to be less relaible than Arrangements B and C.

Additional Particulate Emissions

None of the control devices will contribute to the concentration of any of the combustion generated pollutants, NOx, SOx, and flyash. The SCR and FGD, however, can produce particulate, namely ammonium bisulfate and ammonium sulfate (SCR) and scrubber slurry carry over through the FGD demister. The major contributor to non-flyash particulate is expected to be the FGD, as the SCR production can be controlled by careful control of the ammonia stoichiometry.

Arrangements A and B place the FGD just before the stack. In this position the FGD should additionally control any SCR particulate, but demister carry over is uncontrolled. In Arrangement C, the SCR is just before the stack. The heat exchangers and the SCR should effectively control the FGD particulate by impaction on exposed surfaces; thus with proper SCR control, the additional particulate would be minimal. Based on this discussion, Arrangements A and B should have comparable overall particulate efficiencies somewhat degraded from the baghouse efficiencies; Arrangement C should produce a higher overall efficiency approaching that of the baghouse.

Conclusions

Evaluating the data presented in this report, Arrangement C appears as the optimum configuration due to the lowest overall cost, highest reliability and greatest absolute control efficiency. Particulate and sulfur dioxide are removed upstream of the SCR unit in this configuration, thereby minimizing the possibility of catalyst poisoning and bisulfate/sulfate formation.

Arrangement A would be a more desirable configuration in terms of economics and reliability if the hot ESP was not needed upstream of the SCR unit. Manufacturers

of parallel flow SCR systems claim that their systems can tolerate high dust loadings, therefore, an upstream particulate removal device would not be necessary. The actual necessity or desirability of this device should be determined.

The means of mitigating particulate matter, namely ammonium bisulfate/sulfate formation and/or carry over of scrubber sludge through the demister, potentially generated by the control devices should also be investigated. The concentration of this particulate could place the emission control system out of compliance even though individual particulate control devices may meet prescribed regulations.

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SECTION 1

INTRODUCTION

The Pacific Gas and Electric Company (PG&E) is planning the construction of two 800-megawatt electric generating units, Fossil 1 and 2. The facility will use Western coal with the capability of burning fuel oil and agricultural/municipal wastes.

The California Energy Commission (CEC) is reviewing the PG&E plans within the Notice of Intent (NOI) procedure. Acurex Corporation of Mountain View, California, along with SCS Engineers of Long Beach, California, and CH_2M Hill of Bellevue, Washington, furnished the CEC with information concerning air, water, and solid emissions generated by the combustion of coal and the associated pollution control technologies required to meet emission standards. The information was contained in the Acurex final report entitled "Environmental Control Technologies for a Northern California Coal-Fired Power Plant" (Reference 1).

This supplemental report presents an analysis of those candidate air pollution abatement schemes presented in Section 3.4 of that Final Report. Three basic configurations were identified in that report as workable arrangements. These configurations are analyzed with respect to their capital, operating, and maintenance cost impacts, reliability implications, and overall system removal efficiencies. The components of each arrangement are expected to be the best available control technologies

(BACT) by the 1985-1986 plant startup date. The exception to this may be selective catalytic reduction (SCR) which will be physically installed when it becomes BACT.

1.1 PROCESS EQUIPMENT SCHEMES

The CEC has indicated that the most likely air pollutant control levels and devices to be used will be as follows (Reference 2):

- Particulate -- a control level of 0.005 lbs/MMBtu by the use of a baghouse
- SO_2 -- a control level of 0.07 lbs/MMBtu (based on 0.8 percent sulfur coal and 95 percent control of SO_2) by use of a lime/limestone flue gas desulfurization (FGD) system
- NO_x -- a near term control level of 0.45 lbs/MMBtu by the use of combustion modifications and future control levels of 0.034 lbs/MMBtu by the use of advanced low NO_x burners and selective catalytic reduction

Of the control technologies indicated to meet these emission control levels, some are presently under development, such as selective catalytic reduction (SCR) and state-of-the-art, such as flue gas desulfurization (FGD). Others are existing "best available control technologies" which include electrostatic precipitators (ESP) and fabric filter baghouses (BH). It is anticipated that by the 1985-1986 plant startup dates, FGD will have become BACT for SO_2. Similarly, advanced low NO_x burners and combustion modifications are also expected to be BACT by 1985-1986. SCR will most likely not become BACT until the late 1980's. Actual construction of the plant, therefore, will require leaving adequate space for future installation of this equipment.

Of the many possible schemes, three candidate arrangements have been selected. These arrangements are presented in Figure 1-1 showing only the control equipment. In each of these arrangements, the FGD system is the limestone process because of its advanced development and current use in the utility industry. Another FGD process currently under development is the spray dryer system. Its present status of development will not allow consideration for this project; however, its potential warrants consideration for future projects. A brief discussion of arrangements using a spray dryer FGD system is included in the appendix.

1.2 ANALYSIS APPROACH

Basic information used in the analysis of each component and system was obtained from the Acurex final report. Further background information was obtained from literature sources such as the U.S. Environmental Protection Agency, the U.S. Department of Commerce, and the National Technical Information Service, as well as manufacturers and users.

Data for each component constituting the systems were compiled independent of sequential process arrangement in Section 2. The data, presented in Section 3, discusses the impacts of various design options on cost and reliability which might result from placement of the control equipment. This analysis also includes impacts on operating parameters, independent of system configuration, and special requirements for materials of construction.

Capital, installation, and operating/maintenance cost data are amortized to reflect CEC escalation guidelines as follows (Reference 3):

- Dollar year basis: 1977 and 1985
- Fixed charge rate: (coal) 16 percent

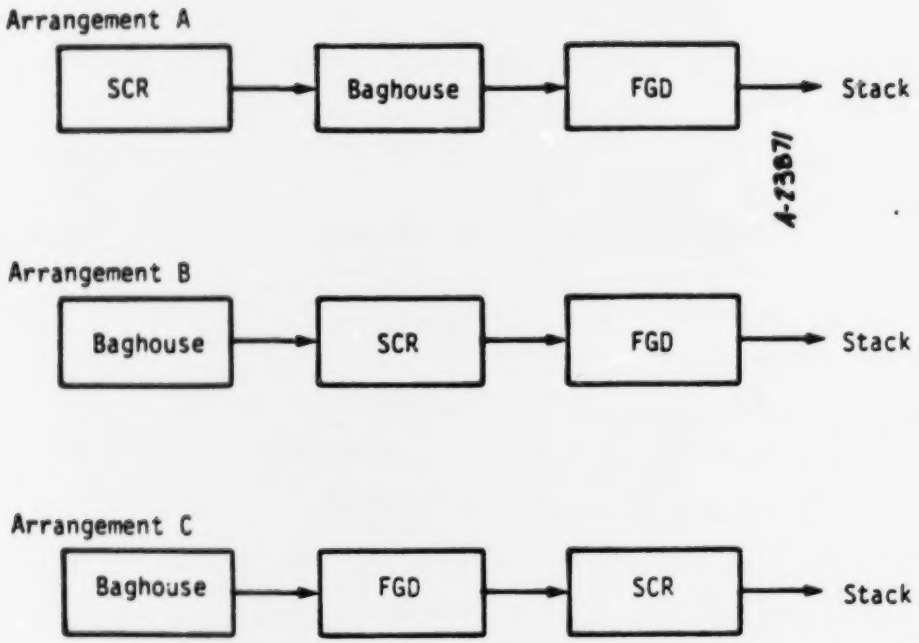

Figure 1-1. Candidate control equipment arrangements.

- Inflation rate: (pollution control) 5.7 percent/year
  (operation and maintenance)
  5.4 percent/year
- Discount rate: 11.5 percent
- Cost of energy: (coal) $0.068/kWh 1985 levelized

When following these guidelines, it should be noted that current economic indicators point to substantially higher rates of escalation.

Compiled capital equipment costs reflect the costs of manufactured equipment and installation. Installation costs include direct (materials) and indirect (engineering, construction and installation labor) at approximately 125 percent of equipment cost. Operating and maintenance costs reflect the costs of operating/maintenance materials, labor, and energy. These costs include annualized capital costs at 10 percent interest, amortized over 10 years.

Energy costs other than coal are amortized to 1985 and are used in operating/maintenance cost calculations. The costs used in this study are as follows for a 5.4 percent inflation rate to 1985:

Electrical power	29 mills/kWh
Steam Energy Cost	36 mills/kWh
Natural gas	2.46 $/MMBtu
Distillate fuel oil	4.00 $/MMBtu
90 percent purity lime delivered	66 $/ton
Soda ash delivered	105 $/ton
Sulfur credit	32 $/ton
Sodium salt cake credit	15 $/ton
Operating labor cost	22 $/hr.

Reliability analysis is qualitatively addressed both on systems and components. Such analysis reflects the state of development of BACT at present and in 1985, and past performance of proven and accepted control techniques. Absolute control efficiencies are addressed with respect to each system arrangement and component device. The absolute removal efficiency of one device may have substantial impact on the efficiency of ensuing downstream devices. It should be noted that, for coal-fired boilers, there is no actual experience reported for any configuration involving combustion modifications, flue gas desulfurization (FGD), SCR, and particulate removal to the levels of pollutant emissions required. Therefore, effects on reliability and efficiency can only be described qualitatively.

Finally, a summary of the equipment arrangements is presented (Section 4) with respect to capital costs, operating costs, efficiency and reliability trade-offs.

SECTION 2

INDIVIDUAL COMPONENT ANALYSIS

The purpose of the following discussion is to present a detailed discussion of auxiliary equipment and component design variations potentially required by a particular arrangement. Similar discussions of the primary control devices are contained in Section 3 of the Acurex final report.

2.1 STACK GAS REHEAT AND AUXILIARY HEATERS

Stack gas reheat is required following the FGD system to avoid downstream condensation and corrosion (and potentially the need for a stack liner), to reduce plume visability, and to enhance plume rise and the dispersion of residual pollutants into the atmosphere. Reheat methods used are in-line reheat (using steam in heat exchangers contacting the flue gas), indirect reheat (using steam in heat exchangers to heat air which is then mixed with the flue gas), or direct combustion reheat (using fossil fuel combustion gases which are mixed with the flue gas).

Auxiliary heaters are needed in some configurations to reheat the flue gas to the SCR process temperature of 750^{o}F. Direct combustion would most likely be the method of choice due to the required gas temperature.

In-line reheaters have encountered severe corrosion problems due to the chloride concentrations in the FGD scrubbing solution. In addition,

material escaping the mist eliminator can cause impact scaling of the heat exchanger surfaces. Indirect hot air heaters are typically free from corrosion and plugging but add large volumes of additional gas which must be handled Direct combustion heaters have had flame instability problems as well as corrosion problems but have low capital investment requirements. Operating costs for fossil fuel are high. The cost of high pressure/high temperature steam is based on the cost to replace the lost capacity.

In-line reheat with steam is the most common method used in the U.S. and is included in arrangements A and B. It is installed following mist elimination. The capital cost, as discussed in the final report (Reference 1), is based on the use of 316LSS or higher alloy steels since 304SS has proven unsatisfactory in many operating systems. Soot blowing equipment may be required to control scale formation. Indirect hot air reheat avoids the corrosion/plugging problems and requires less maintenance, but the additional fan, duct work, and controls requires approximately 70 percent more capital and increases utility steam consumption by 50 percent (Reference 4).

Direct combustion reheat requires careful start-up to avoid thermal shock and cracking of the refractory lining. The capital cost of direct combustion reheat is approximately 30 to 35 percent that of in-line and 15 to 20 percent that of indirect reheat systems. Operating and maintenance costs are typically 200 percent that of in-line and 150 percent that of indirect reheat systems (Reference 4).

2.2 COMBUSTION AIR PREHEATER

The air preheater is typically an integral part of the utility boiler system. Arrangement A requires the removal of approximately

95 percent of all particulate prior to the air preheater. Though particulate is removed, no change in materials of construction is anticipated. However, the overall reliability of the boiler system is enhanced and maintenance can be reduced by up to 20 percent (Reference 1).

2.3 BAGHOUSES

The basic difference between the arrangements is in the concentration of dust being controlled by the baghouse. Since the upstream hot ESP collects 95 percent of the particulate in arrangement A, the net particulate collected in the baghouse is reduced by a factor of 20 compared to arrangements B and C. Bag life, however, is not expected to be substantially prolonged due to the size of particles collected and their abrasive quality (Reference 1).

The air to cloth ratio for pulse cleaning baghouses is 2 to 3 times greater than for reverse air and mechanical shaker type baghouses. However, this application is too large for a pulse baghouse. A shaker or a reverse air unit would be appropriate, and the capital costs are approximately the same for each.

A final consideration is bag type. Fiberglass bags are generally used in reverse air baghouses, with Nomex and other more durable bags in pulse clean or mechanical shaker type baghouses. Nomex cloth costs approximately twice as much as fiberglass.

2.4 FLUE GAS DESULFURIZATION (FGD)

The FGD system, as discussed in the following section, requires upstream particulate control. This is due to the operating history of alkaline ash/lime or limestone systems. Those systems which simultaneously remove ash and SO_2 have reduced reactant requirements but also have increased operating and maintenance requirements and a

catalyst. However the down flow arrangement does require extra duct work, and a unit operating at full particulate dust load will also require a flyash removal hopper.

SECTION 3

SYSTEMS ANALYSIS

Three air pollution abatement schemes are being considered for implementation on Fossil 1 and 2. These schemes include the most likely equipment components applicable to coal-fired electric utilities and combine BACT and technology anticipated to be commercially available by 1985. In this section, each of the three system arrangements are analyzed qualitatively on the basis of reliability, absolute control efficiency and capital, and operating and maintenance costs.

Combustion modifications are included in all three arrangements. Advanced low NO_x burners will have little effect on SO_2 concentrations exiting the boiler. However, a slight increase in particle size distribution may result. This may increase the effectiveness of particulate control devices. At present, no quantitative data is available to confirm these suppositions. Therefore, primary concern will be concentrated on control devices following the boiler.

3.1 ARRANGEMENT A

3.1.1 Description

Arrangement A is schematically depicted in Figure 3-1. The flue gas temperature of 750°F leaving the boiler coincides with the required SCR operating temperature. However, particulate blinding effects on the catalyst bed may decrease effectiveness and reliability of the SCR and

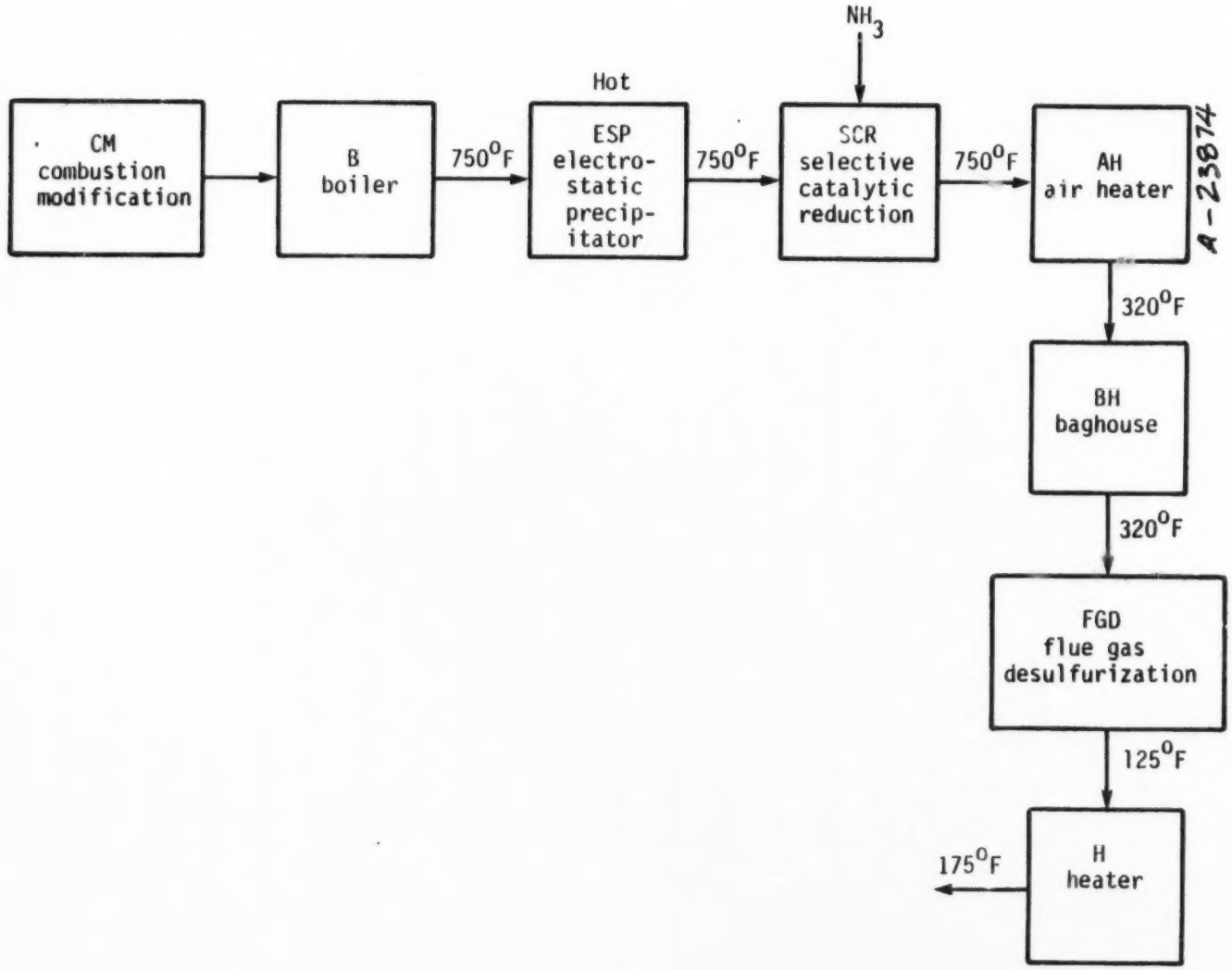

Figure 3-1. Arrangement A -- selective catalytic reduction with electrostatic precipitator, baghouse and flue gas desulfurization.

increase maintenance requirements. A low efficiency, hot ESP may precede the SCR to remove a substantial amount of particulate.

The combustion air preheater following the SCR removes sensible heat from the flue gas. This lowers the flue gas temperature to approximately 320°F before entering a reverse air baghouse fitted with fiberglass bags. The flue gas then goes to the FGD system. The saturated flue gas is reheated in an in-line heat exchanger and is discharged to the stack.

3.1.2 Operational Considerations

The primary attractiveness of arrangement A is that flue gas leaving the boiler's economizer is at the same temperature as the required operating temperature of the SCR process. However, the relatively high particulate grain loading of the flue gas (3 gr/scf) (Reference 1) may adversely affect the catalyst bed. Manufacturers of parallel flow, SCR systems claim that their systems can tolerate high grain loadings; however, these systems have yet to be demonstrated on coal-fired units. Quite possibly the problem of catalyst blinding may be significantly reduced but catalyst errosive should still be a concern. Erosion of the catalyst reduces catalyst life, which is significant due to the large replacement cost. To alleviate this potential problem, a low efficiency (95 percent) hot ESP has been included to remove the bulk of particulate effluents. More experience is required to determine the actual necessity or desirability of this device.

A 95 percent efficient "hot" ESP would effectively remove the majority of particles larger than 2 micrometers (quantitative efficiencies with respect to particle sizes vary). Such a hot ESP would be a dry single stage, flat plate configuration. By operating at a reduced

efficiency (compared to a typical, high efficiency of 99.9 percent), the ESP specific collection area (SCA) would be substantially reduced. However, at elevated gas temperatures, the gas volume increases. Therefore, the size and cost of a hot ESP can be expected to be close to that of a cold, high efficiency ESP. Operating hot, the properties of flyash resistivity are favorably enhanced. In the case of Western coal at 3 percent sulfur, idealized particle resistivity is minimized at temperatures in excess of 400°F. For the case under consideration, the high operating temperature for the ESP (750°F) may minimize the resistivity of the ash which would otherwise be expected for a 0.8 percent sulfur coal. Actual resistivity data with respect to temperature must be compiled before the physical size of the hot ESP can be established. Based on a flue gas particulate loading of 3 gr/scf, 27 tons/hr of flyash are emitted from the boiler. At an ESP efficiency of 95 percent, the hot ESP would produce 25.7 tons/hr of dry flyash for solid waste disposal. Grain loading is substantially reduced before entering the SCR. Due to the low pressure drop (1 in. W.G.) of ESPs, induced draft fan power is not substantially increased and, with the fan placed between the precipitator and the SCR, reduced particulate abrasion in the fan may be realized.

The temperature of the flue gas entering the combustion air preheater is above the sulfuric acid condensation temperature and the reduced grain loading (approximately 0.15 gr/scf) reduces the effects of mechanical particulate erosion on the heat exchanger tubes. Depending on design, metal temperatures at 320°F are well above minimum recommended levels (approximately 200°F) to avoid corrosive attack. Capital costs are reduced by avoiding corrosion resistant stainless steels as a construction material.

With the use of injected NH_3 in the SCR process and any unreacted NH_3 downstream, the formation of ammonium bisulfate (NH_4HSO_4) and ammonium sulfate $(NH_4)_2SO_4$ is possible. Unreacted NH_3 reacts with SO_3 and H_2O to form these compounds at low flue gas temperatures. Ammonium bisulfate condenses at 410°F and may deposit on heat exchanger surfaces. High pressure water washing will remove such deposits but is not a desirable maintenance step. Formation of these compounds may also hamper baghouse operation. Operating the air preheater at a higher outlet temperature (410°F) may alleviate fouling, but is too high a process temperature for reliable baghouse operation. Successful system implementation using an SCR/NH_3 NO_x abatement system is best obtained by the careful monitoring of ammonia to ideal stoichiometric ratios to minimize bisulfate and sulfate formation.

A flue gas temperature of 320°F or lower allows the use of fiberglass bags in the baghouse. Though more economical than Nomex or similar textiles, the fiberglass bags are fragile. Reverse air cleaning must be used to minimize mechanical damage to the dust laden or caked bags. Consequently, an air to cloth ratio of 2 should be used resulting in net cloth area, size and cost greater than those of pulse or mechanical shaker type baghouses with higher air to cloth ratios (up to 15). Due to the reduced particulate loading as a result of the use of a hot ESP, total baghouse cloth area is reduced. Based on a 95 percent efficient ESP, 1.3 tons/hr of flyash are removed by the baghouse. For an anticipated efficiency of 99.92 percent, combined with a very low grain loading, implementation of a fabric filter collector is feasible both from operational and economic viewpoints. Ammonium sulfate compounds are "sticky" at 320°F and may adhere in the bag weave. Combined with

flyash, silica-based particulate, the result could be the excessive caking of the bags. Such an occurrence would render reverse air cleaning of the fiberglass bags ineffective. Pulse cleaning or shaking methods used with Nomex bags would be required over that of fiberglass.

By locating the limestone FGD process as in arrangement A, flue gas enters the scrubber following particulate removal. With virtually no flyash, the problem of wet-dry interface disposition of solids is alleviated.

Mist eliminators are designed to remove condensed water and slurry carryover from the scrubber. The effectiveness of the demister will affect the entrained particulate exiting the stack. Stack gas reheat is used to add heat to the saturated gas stream to improve bouyancy, reduce downstream condensation and corrosion, and reduce plume visibility and acid rain problems. A stack gas temperature of $175^{o}F$ has been assumed in this analysis. Such a heater would be designed for in-line application with scaling, erosion and corrosion resistant materials. Flue gas leaving the air heater at $175^{o}F$ will minimize downstream acid condensation problems. By proper design and insulation of the ductwork and stack, the need for a wet liner in the stack may be eliminated.

3.1.3 Absolute Removal Efficiencies and Reliability

NO_x, SO_x, and flyash are combustion products. Therefore, the arrangement of control equipment following the combustion zones should not contribute to the concentration of any of these pollutants in the effluent gas stream. Furthermore, the outlet concentrations of these pollutants from the stack will be a function of the removal efficiencies of the individual control devices.

Particulate matter other than flyash can be generated by the control devices. As described above, this particulate can be the reaction products ammonium bisulfate and ammonium sulfate and/or carryover of scrubber slurry through the demister.

In arrangement A the baghouse controls any ammonium sulfate/bisulfate particulate generated. However, the flue gas desulfurization system is not followed by a particulate control device other than the FGD demister. The demister, as the name implies, limits the mist -- particulate as well as liquid droplet emissions -- from the scrubber to the atmosphere. The mist elimination results in prevention of corrosion and/or erosion of the exhaust stack lining, the induced draft fan, and in-line stack gas reheater tubes. The predominant mechanism of mist collection is inertial impaction. The predominant re-entrainment mechanisms are high-velocity stripping and drag flow of collected particles off the mist-eliminator surfaces. The outlet particulate concentration may, therefore, be affected by the operation of this mist eliminator. Literature sources do not have information which specifically quantified demister efficiency and particulate concentration. Therefore, individual particulate control devices may meet the prescribed regulations, but the system could be out of compliance.

3.1.4 Capital and Operating Costs

The estimated system and operating costs are tabulated in Table 3-1. Total capital cost for arrangement A is estimated to be 160×10^6 dollars in 1985. Total operating cost is estimated to be 33×10^6 dollars in 1985. These costs exclude the cost of combustion modification NO_x controls, which are used for all arrangements.

TABLE 3-1. TABULATED FLOW AND EFFICIENCY PERFORMANCE PARAMETERS ARRANGEMENT A

	Hot ESP	SCR	Baghouse	FGD	Stack Heater
Method	Hot Plate Type	Parallel Flow	Reverse Air	Limestone	Steam or Fuel Oil
Flow Data Flue Gas	3.87×10^6 acfm at 750°F	3.87×10^6 acfm at 750°F	2.49×10^6 acfm at 320°F	1.86×10^6 acfm at 125°F	2.02×10^6 acfm at 175°F
Design Efficiency	95% Flyash	90% NO_x	99.92% Flyash	95% SO_2	--
Device Pressure Drop	ΔP = 1" W.G.	ΔP = 9" W.G.	ΔP = 6" W.G.	ΔP = 9" W.G.	ΔP = 1" W.G.
Energy Req. Conveyance Heating	0.39 MW --	3.5 MW --	2.3 MW --	3.5 MW --	0.34 MW 30 MW
Energy Costs Conveyance Heating	\$78 x 10^3/yr	\$700 x 10^3/yr	\$470 x 10^3/yr	\$78 x 10^3/yr	-- \$6.6 x 10^6/yr
Operating Costs 1985	\$4.9 x 10^6/yr	\$6.2 x 10^6/yr	\$2.2 x 10^6/yr	\$13 x 10^6/yr	\$6.6 x 10^6/yr
Capital Costs 1985	\$31 x 10^6 \$39/kW	\$39 x 10^6 \$48/kW	\$22 x 10^6 \$27/kW	\$71 x 10^6 \$88/kW	\$0.28 x 10^6 \$0.27/kW
NOTES	Installed Insulated	For NH_3/NO = 1 15 yr Reactor Life	Air/Cloth = 2 Installed Insulated Fiberglass Bags		

3.2 ARRANGEMENT B

3.2.1 Description

This arrangement is schematically depicted in Figure 3-2. Flue gas leaves the boiler and enters the combustion air preheater. The outlet gas temperature is reduced to 320°F, lc enough for baghouse operation. Particulate is removed in the baghouse. To attain the proper SCR operating temperature of 750°F, a heat exchanger and supplemental heater are utilized. Sensible heat is removed from the flue gas leaving the SCR by use of a second heat exchanger. The supplemental heater is of the in-line heat type, but could be by direct fossil fuel combustion. Flue gas leaving the second heat exchanger (following the SCR system) then enters the FGD system. The flue gas is scrubbed of SO_2 and leaves the FGD system at 125°F. Finally, the flue gas is reheated to 175°F and emitted to the atmosphere.

3.2.2 Operational Considerations

In contrast to arrangement A, this arrangement requires three heat transfer steps. This arrangement is more energy intensive than arrangement A as the various heaters require a heat transfer fluid for the recovery of energy from the SCR effluent and cycle steam or auxiliary fossil fuel.

Flue gas leaving the boiler econimizer enters the combustion air preheater. Again, with proper outlet temperature above 200°F, sulfur attack on air heater components is minimized. However, the relatively high particulate loading (3 gr/scf) may cause mechanical erosion of the preheater surfaces.

Compared to arrangement A, the baghouse placement in arrangement B is advantageous. Upstream of the SCR process, baghouse performance is not

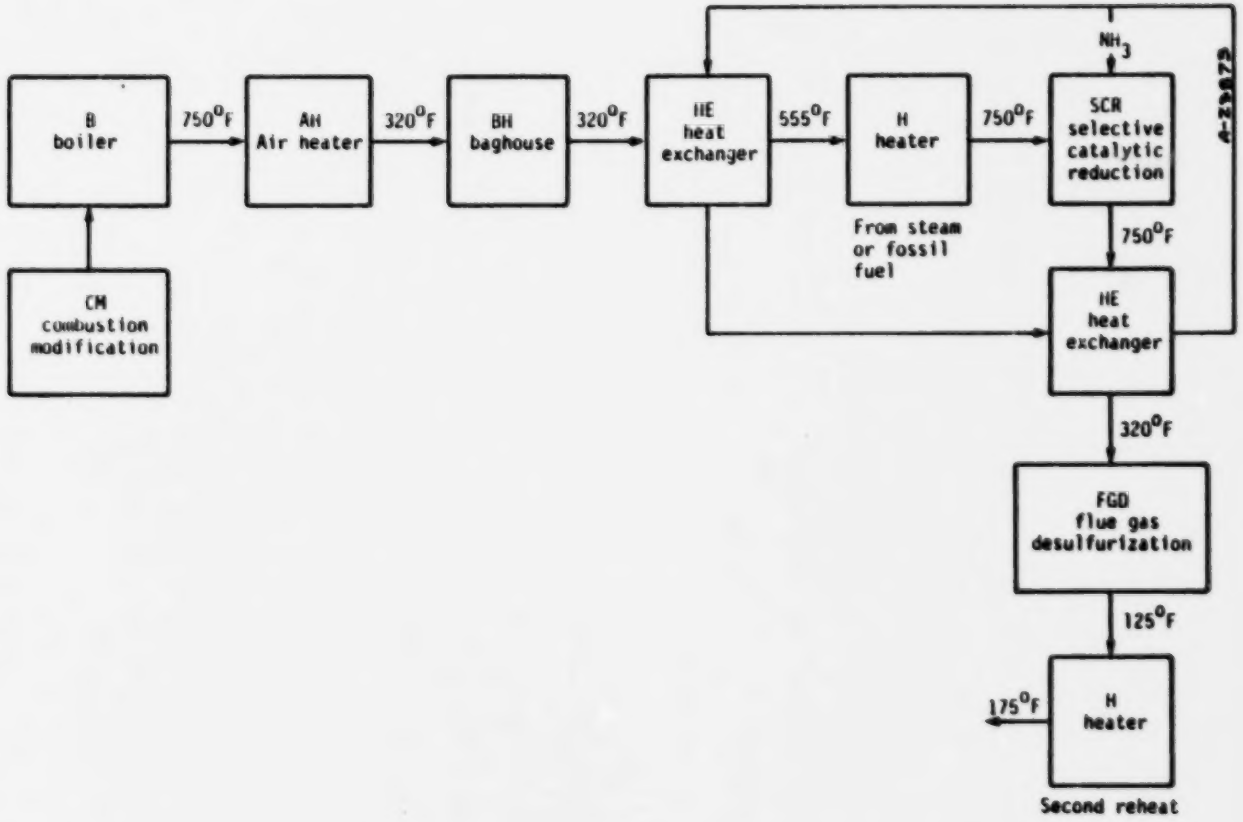

Figure 3-2. Arrangement B -- Selective catalytic reduction with baghouse, flue gas desulfurization and flue gas reheat.

affected by the formation of ammonium sulfate/bisulfate compounds and the potential adherence on the bag surface. Particulate is removed at 99.92 percent or better efficiencies. Baghouse size is substantially larger as the grain loading is high. Baghouse operation in this configuration has been demonstrated. For the sake of comparison, a reverse air baghouse with fiberglass bags is implemented; however, a shaker baghouse can be used. The total capital cost and operating and maintenance cost for each system is approximately equal.

By the implementation of a primary heat exchanger, flue gas can be reheated from 320°F to 555°F. By the use of a superheated steam, in-line heater, flue gas is reheated to 750°F which is the process temperature required by the SCR process. The injection of ammonia for NO_x control can lead to the generation of ammonium bisulfate (NH_4HSO_4) and ammonium sulfate $(NH_4)_2SO_4$. Liquid ammonium bisulfate forms at 410°F and will deposit on the secondary heat exchanger components with wall temperatures less than 410°F. This may increase scaling and corrosion-erosion and consequently increase the maintenance requirements of that component. However, heat exchanger fouling due to flyash is minimized as the flue gas is essentially flyash free downstream of the baghouse. Moreover, the effects of particulate erosion and blinding in the reactor catalyst bed are minimized.

The flue gas enters the FGD process at 320°F. As in the case of arrangement A, a limestone process at a 95 percent efficiency is considered. The FGD system is expected to control any ammonium sulfate/bisulfate generated. As in the case of arrangement A, the flue gas is demisted, reheated to 175°F, and discharged to the atmosphere.

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3.2.3 Absolute Removal Efficiencies and Reliability

As in arrangement A, the NO_x, SO_x, and flyash combustion products are not affected by control device location. In addition, the last control device is the FGD process, resulting in a potential particulate generation problem, depending on demister operation.

Compared to arrangement A, the reliability of the air preheater may be reduced due to flyash erosion or plugging. Baghouses have been successfully operated at particulate levels of 3 gr/scf; therefore, no abnormal reliabilities for this arrangement are expected. In the absence of particulate, the SCR catalyst bed is not eroded nor blinded, thereby reducing maintenance requirements and enhancing reliability. The balance of the system is expected to operate similarly to arrangement A.

3.2.4 Capital and Operating Costs

The estimated system and operating costs are tabulated in Table 3-2. Total capital costs for arrangement B is 140×10^6 dollars in 1985. Total operating cost is 52×10^6 dollars in 1985. These costs exclud the cost of combustion modification NO_xcontrol.

3.3 ARRANGEMENT C

3.3.1 Description

This arrangement is schematically depicted in Figure 3-3. As in arrangement B, flue gas leaving the boiler at 750°F enters the primary combustion air heater. The outlet temperature is 320°F for baghouse operation. Fiberglass bags and reverse air cleaning remove particulate at 99.92 percent efficiency prior to the FGD process. Unlike arrangement B, flue gas is not reheated but enters the FGD system. After the wet scrubbing for the removal of SO_2, flue gas at 125°F enters a heat exchanger and exits at 555°F. To reach the SCR process temperature a

TABLE 3-2. TABULATED FLOW AND EFFICIENCY PERFORMANCE PARAMETERS ARRANGEMENT B

	Baghouse	Heat Exchanger	Heater	SCR	Heat Exchanger	FGD	Heater
Method	Reverse Air	Plate or Tube	Steam or Fuel Oil	Parallel Flow	Plate or Tube	Limestone	Steam or Fuel Oil
Flow Data Flue Gas	2.48×10^6 acfm at 320°F	3.23×10^6 acfm at 555°F	3.87×10^6 acfm at 750°F	3.87×10^6 acfm at 750°F	2.49×10^6 acfm at 320°F	1.86×10^6 acfm at 125°F	2.02×10^6 acfm at 175°F
Design Efficiency	99.92% Flyash	--	--	90% NO_x	--	95% SO_2	--
Device Pressure Drop	ΔP = 6" W.G.	ΔP = 5" W.G.	ΔP = 1" W.G.	ΔP = 9" W.G.	ΔP = 5" W.G.	ΔP = 9" W.G.	ΔP = 1" W.G.
Energy Req. Conveyance Heating	2.3 MW --	2 MW --	0.39 MW 117 MW	3.4 MW --	2 MW --	3.4 MW --	0.34 MW
Energy Costs Conveyance Heating	\$470 x 10^3/yr --	\$390 x 10^3/yr --	\$78 x 10^3/yr \$26 x 10^6/yr	\$700 x 10^3/yr --	\$390 x 10^3/yr --	\$700 x 10^3/yr --	\$78 x 10^3/yr \$6.6 x 10^6/yr
Operating Costs 1985	\$2.2 x 10^6/yr	\$0.6 x 10^6/yr	\$26 x 10^6/yr	\$6.2 x 10^6/yr	\$0.6 x 10^6/yr	\$13 x 10^6/yr	\$6.6 x 10^6/yr
Capital Costs 1985	\$22 x 10^6 \$21/kW	\$3.7 x 10^6 \$4.70/kW	\$0.29 x 10^6 \$0.36/KW	\$39 x 10^6 \$48/kW	\$3.7 x 10^6 \$4.70/kW	\$70 x 10^6 \$88/kW	\$0.2 x 10^6 \$0.27/kW
NOTES:	Installed Insulated Fiberglass Bags	Reference 5	Steam Energy Cost		Reference 5		Steam Energy Cost

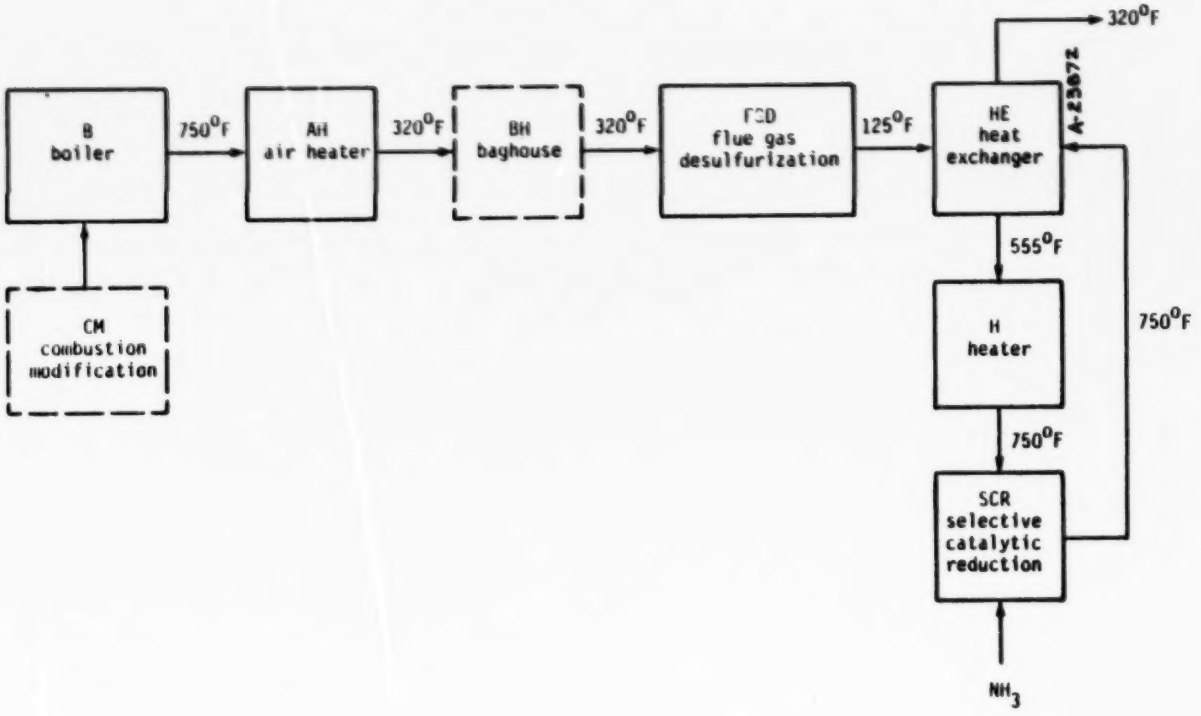

Figure 3-3. Arrangement C -- Flue gas desulfurization with baghouse, selective catalytic reduction and flue gas recirculation.

superheated steam, in-line, is used to reheat the flue gases to 750°F. The gas exiting the SCR passes to the heat exchanger downstream of the FGD for recovery of the sensible heat. The flue gas exits at 320°F and needs no further reheating prior to atmospheric release.

3.3.2 Operational Considerations

As in arrangement B, the flue gas temperature is reduced in the air preheater, followed by particulate control in the baghouse. The gas then flows to the FGD system for SO_2 control. See the previous Section 3.2 for a discussion of the system to this point.

Flue gas, scrubbed of SO_2, leaves the FGD system as saturated gas at 125°F. Though the SO_2 content has been substantially reduced, the process temperature and the presence of residual sulfur compounds represent severe corrosion problems for equipment downstream of the FGD system. In this arrangement, a heat exchanger utilizing recirculated flue gas from the SCR heats the flue gas from 125°F to 555°F. Materials of construction in the heat exchanger must be corrosion resistant, high chromium and/or nickel steels representing a 250 percent increase in cost over that of standard construction materials. A reduction in heat exchanger efficiency may occur due to demister inefficiency and may therefore, require substantial routine cleaning.

The SCR process will remove NO_x at 90 percent efficiency. Problems associated with flyash poisoning the reactor have been essentially eliminated by the upstream baghouse and FGD process. However, particulate escaping the mist eliminator may interfere with reactor operation due to its "sticky" characteristics. The primary heat exchanger (following the SCR) has an exit temperature of 320°F which allows the possible formation and deposition of ammonium sulfate/bisulfate on heat

exchanger components causing degradation in efficiency and corrosion. As discussed earlier, corrosion resistant materials of construction must be used in the heat exchanger. Penalties for efficiency degradation consists of additional fuel consumption to reach SCR process temperature requirements and increased corrosion on primary heat exchanger components.

Auxiliary stack gas reheat is not required due to the high outlet gas temperature.

3.3.3 Absolute Removal Efficiencies and Reliability

Once again, this arrangement will not generate SO_x, NO_x, or flyash. However, both the FGD system and the SCR can potentially contribute to outlet particulate loadings. Unlike arrangement B, the SCR now follows the FGD system. Any particulate escaping the mist eliminator must pass through two heat exchanger systems and the SCR. The amount of surface area for deposition of particulate is far greater than in the other arrangements. This is expected to substantially reduce the calcium salt/SO_3 mist entrained in the gas stream. The calcium salt/SO_3 mist may reach the SCR catalyst and reduce the effective life of the bed. The SCR generated particulates will pass through one heat exchanger but this represents less in terms of surface area than arrangement B; therefore, particulate capture will be reduced. Since the FGD system is expected to be the major contributor of additional particulate matter (i.e., non-flyash), arrangement C should have the highest absolute particulate efficiency.

3.3.4 Capital and Operating Costs

The estimated system and operating costs are tabulated in Table 3-3. Total capital cost for arrangement C is estimated to be 130×10^6 dollars in 1985. Total operating cost is estimated to be 48×10^6 dollars in 1985. These costs exclude the cost of combustion modification NO_x control.

TABLE 3-3. TABULATED FLOW AND EFFICIENCY PERFORMANCE PARAMETERS ARRANGEMENT C

	Baghouse	FGD	Heat Exchanger No. 1	Heater	SCR	Heat Exchanger Hotside
Method	Reverse Air	Limestone	Plate or Tube/ Cold Side	Steam or Fuel Oil	Parallel Flow	Hot Side
Flow data Flue gas	2.48×10^6 acfm at 320°F	1.86×10^6 acfm at 125°F	3.23×10^6 acfm at 555°F	3.87×10^6 acfm at 750°F	3.87×10^6 acfm at 750°F	2.71×10^6 acfm at 320°F
Design Efficiency	99.92% Flyash	95% SO_2	--	--	90% NO_x	--
Device Pressure Drop	ΔP = 6" W.G.	ΔP = 9" W.G.	ΔP = 5" W.G.	ΔP = 1" W.G.	ΔP = 9" W.G.	ΔP = 5" W.G.
Energy Req. Conveyance Heating	2.3 MW --	3.5 MW --	2 MW --	0.39 MW 117 MW	3.4 MW --	2 MW --
Energy Costs Conveyance Heating	\$470 x 10^3/yr --	\$700 x 10^3/yr --	\$390 x 10^3/yr --	\$78 x 10^3/yr \$26 x 10^6/yr	\$700 x 10^3/yr --	\$78 x 10^3/yr --
Operating Costs 1985	\$2.2 x 10^6/yr	\$13 x 10^6/yr	\$0.46 x 10^6/yr	\$26 x 10^6/yr	\$6.2 x 10^6/yr	Incl. in HE # 1
Capital Costs 1985	\$22 x 10^6 \$27/kW	\$71 x 10^6 \$80-90/kW	\$2.8 x 10^6 \$3.60/kW	\$0.2 x 10^6 \$0.27/kW	\$39 x 10^6 \$48/kW	Incl. in HE # 1
	Installed Insulator		Stainless Steel Construction	Steam Energy Cost		

SECTION 4

SUMMARY

A summary of the three arrangements is shown in Table 4-1. Arrangement A has the highest capital cost of the three arrangements by 18 percent at $200 per kilowatt. However, operating and maintenance costs are lower by over 32 percent at 6.7 mills per kilowatt-hour in 1985. Arrangements B and C have lower capital costs as no ESP is required. The higher operating and maintenance costs result from additional flue gas reheating. The capital costs of Arrangements B and C are similar at $170 per kilowatt.

This comparison of costs must be viewed in terms of the accuracy of the cost estimates. The estimates used are considered to be mid-range with a variance of ±35 percent in general. Consequently, the cost differences for the arrangements which are well within the variance may not be significant. Also, site specific considerations have not been included. For example, arrangement A, which requires an additional major unit (ESP), may present space allocation problems. Also, since the SCR is to be incorporated later, its relative location may have additional cost implications.

There is no actual experience reported for any control configuration involving combustion modifications, flue gas desulfurization, selective catalytic reduction and baghouses and/or hot ESP. Further, two of these

TABLE 4-1. SUMMARY OF VARIOUS AIR POLLUTION ABATEMENT SCHEMES

Description	Arrangement A	Arrangement B	Arrangement C
Abatement System Capital Costs in 1985 (excludes air heater -- part of boiler package)	$160 x 10^6	$140 x 10^6	$140 x 10^6
Capital Cost per Kilowatt	$200/kW	$170/kW	$170/kW
Operating and Maintenance Costs in 1985 (includes air heater)	$33 x 10^6/yr	$52 x 10^6/yr	$48 x 10^6/yr
Cost per Kilowatt Hr	6.7 mills/kWh	11 mills/kWh	9.8 mills/kWh
System Reliability	Fair	Good	Good
Absolute System Efficiency	Fair	Fair	Good

control technologies, SCR and hot ESP, are still in the development stage which seriously limits confidence in reliability estimates. Finally, the control efficiencies required for this project are very stringent, which limits the usefulness of past history even for BACT devices. Consequently, effects on reliability and overall efficiency can only be described qualitatively.

Arrangement A requires two devices (SCR and hot ESP) that are in the development stage whereas arrangements B and C require only one (SCR). Other reliability problems, such as heat exchanger corrosion, can be designed around (at higher cost) to assure high reliability. As more experience is gained with FGD operation, its reliability is steadily improving, probably exceeding 90 percent currently. Baghouses, because of their modular design, have very high reliability approaching 100 percent. Therefore, the limiting reliability will probably be the SCR or hot ESP; based on this, arrangement A should be viewed as less reliable than arrangements B and C. No significant synergistic reliability problems have been identified.

None of the control devices will contribute to the concentration of any of the combustion generated pollutants, NO_x, SO_x, and flyash. The SCR and FGD, however, can produce particulate, namely ammonium bisulfate and ammonium sulfate (SCR) and scrubber slurry carry over through the FGD demister. The overall efficiency for NO_x and SO_x removal, therefore, is unaffected by control device arrangement, but the final particulate concentration can be affected.

The major contributor to non-flyash particulate is the FGD as the SCR production can be controlled by careful control of the ammonia stoichiometry. Arrangements A and B place the FGD just before the stack.

In this position, the FGD should additionally control any SCR particulate, but demister carry over is uncontrolled. In arrangement C the SCR is just before the stack. The heat exchangers and the SCR should effectively control the FGD particulate by impaction on exposed surfaces; thus, with proper SCR control, the additional particulate would be minimal. No quantitative information on demister efficiency is reported to allow one to estimate the severity of the problem. Based on this discussion, arrangements A and B should have comparable overall particulate efficiencies somewhat degraded from the baghouse efficiency; arrangement C should produce a higher overall efficiency approaching that of the baghouse.

The control devices discussed for the control of NO_x, SO_x and particulate will have virtually no effect on the carbon monoxide (CO) emissions due to its low solubility. Hydrocarbons, however, may be reduced either by dissolution in the FGD or condensation on particulate that is subsequently removed. There is no information to quantify this effect.

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APPENDIX A

FGD/SPRAY DRYER DISCUSSION

Lime/limestone FGD systems were focused on in this study because of their relatively advanced development and extensive use in the utility industry. Spray dryer FGD processes, however, are under development and warrant consideration for future projects depending on their progress. To date, no full scale, dry scrubbing systems of this type have been constructed for the utility industry. Two full scale systems have been sold, with startup scheduled for the early 1980's. At least seven vendors (generally spray dryer manufacturers and dust collector manufacturers under joint agreement) have process designs which they offer commercially. Generally, each of the manufacturers has significant experience in marketing their systems separately. However, only pilot scale systems combined into a single system have been operated to date in the utility sector.

The method of operation of the spray dryer absorber system is relatively simple. The sorbent solution is atomized into the incoming gas stream to promote mass transfer of the SO_2 to the liquid phase. The volume of water added to the gas phase by evaporation is insufficient to saturate the gas, resulting in a dry reaction product that can be disposed of accordingly.

The spray dryer absorber system can be divided into two basic processing steps:

- Spray dryer absorber
- Dust collector

A flow diagram is shown in Figure A-1. If no SCR were required, but only SO_x and particulate controls applied, the flue gas would pass from the boiler house directly to the spray dryer. The dryer is designed to accept the full dust load emitted from the boiler. The gas enters the spray dryer tangentially through the roof via spinning vanes. This allows for maximum gas/liquid contact, a critical design parameter. The gas contacts the absorption media, then flows to the dust removal equipment. The outlet temperature of the gas going to the stack is controlled at a temperature above the saturation point.

The dust collection equipment can be designed to receive the total mass of particulate and reacted salts from the spray dryer (or a lesser amount if dust can be discharged from the dryer itself). The clean gas then exits to the stack. The particulate matter collected is discharged from the system as a dry solid. The flyash/reaction product mixture can then be disposed of as a landfill material.

The SCR unit can be introduced either upstream of the spray dryer/baghouse combination (similar to arrangement A) or downstream (similar to arrangement C). A significant difference between the spray dryer arrangement A and the lime/limestone case is that, for the former, the baghouse is downstream of all devices. This is feasible due to the "dry" effluent at this point. Therefore, the overall efficiency of particulate collecton is that of the baghouse, i.e., ammonium bisulfate

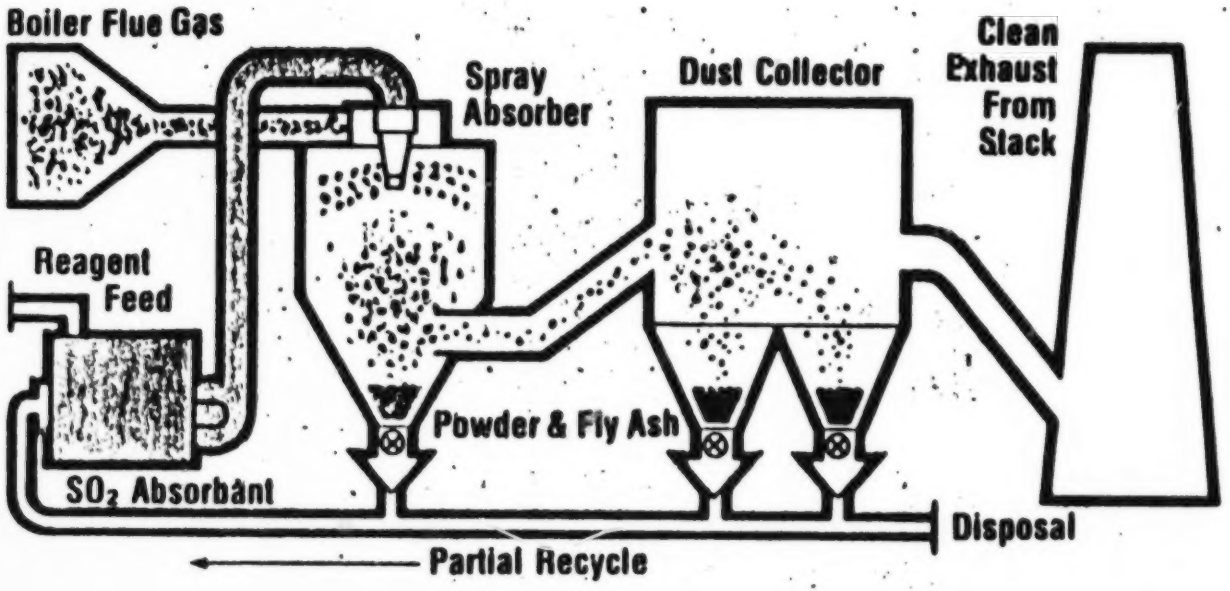

Figure A-1. Spray dryer SO_2 abatement system.

and ammonium sulfate from the SCR will be captured and there is no scrubber mist generated.

The spray dryer/arrangement C has the advantage that the scrubber mist will not be deposited in the downstream heat exchanger (required for the SCR) or on the SCR catalytic surface. Elimination of these problems may improve overall reliability. The emissions of sulfates from the SCR must still be controlled by proper NH_3 stoichiometry.

The spray dryer/FGD process is still under early development and its potential removal efficiency and reliability have not been established. The process does have some inherent advantages, due to the "dry" effluent, in effecting higher overall efficiencies and/or reliabilities.